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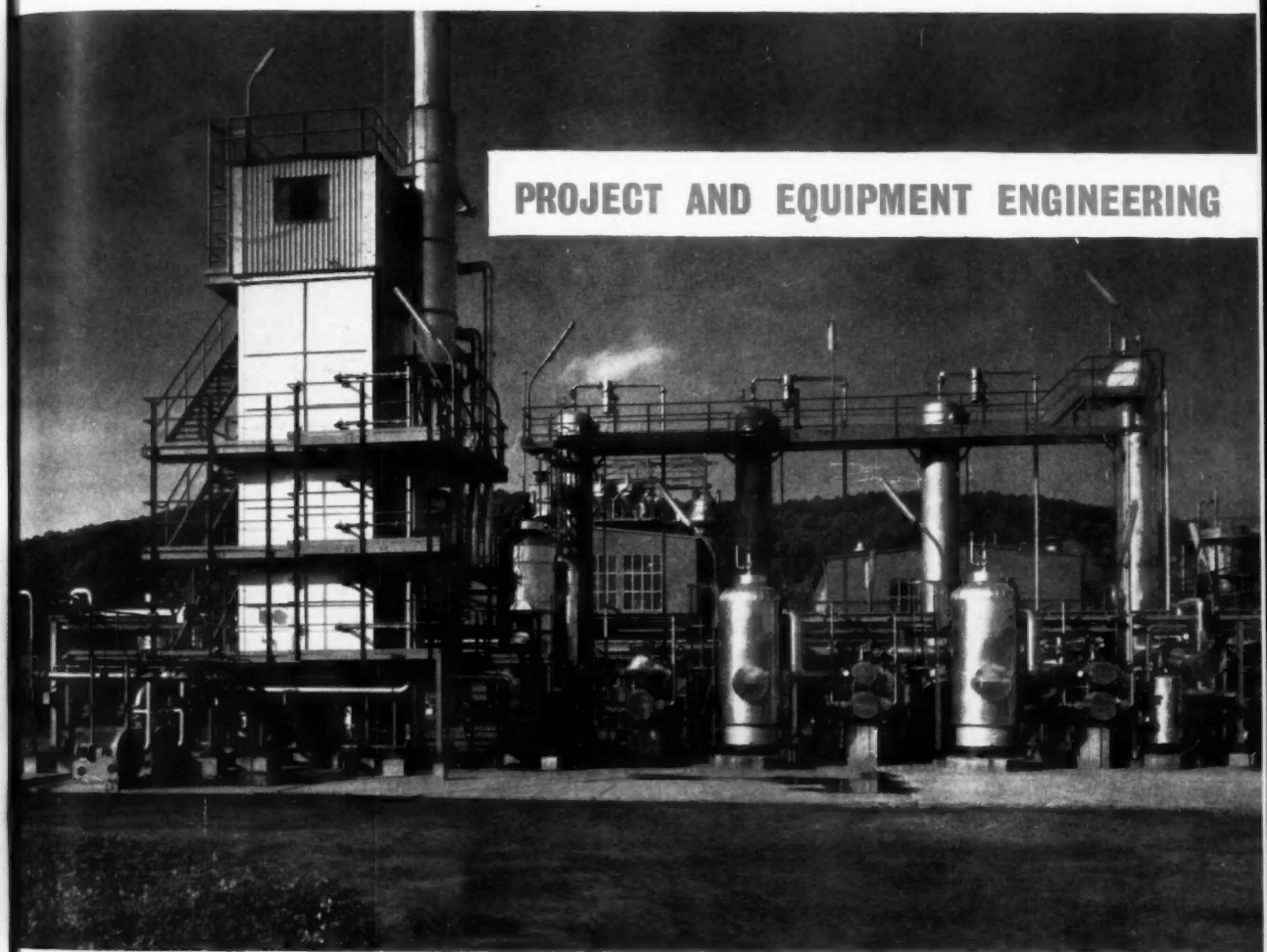

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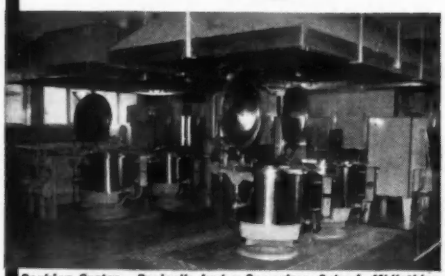

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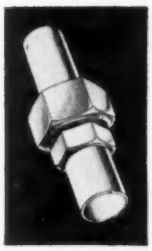

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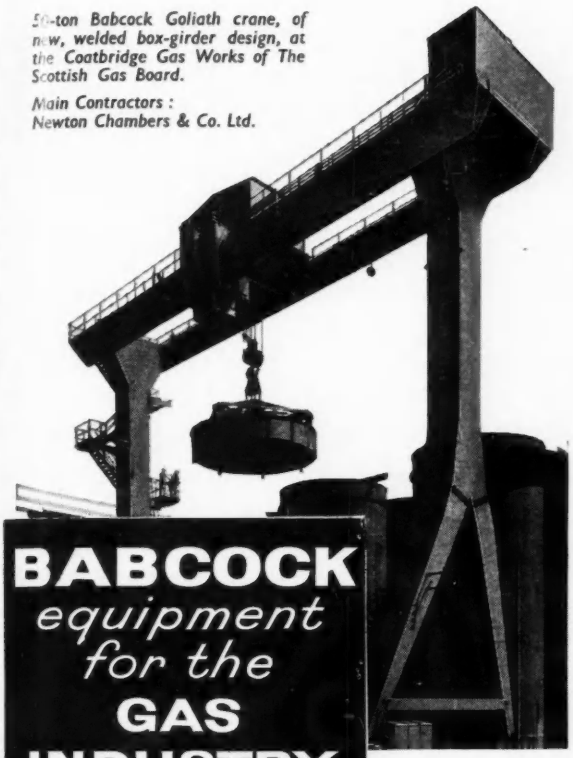

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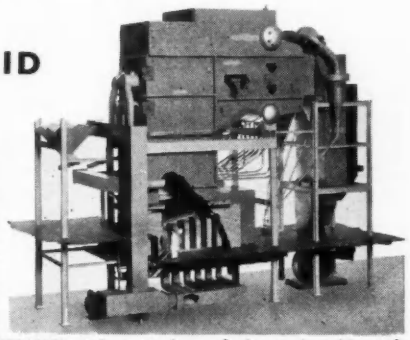

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Editor:
GEOFFREY W. BATTISON

Technical Editor:
CHARLES J. P. DE WINTON, T.D., M.A., M.Inst.Gas E., M.Inst.F.

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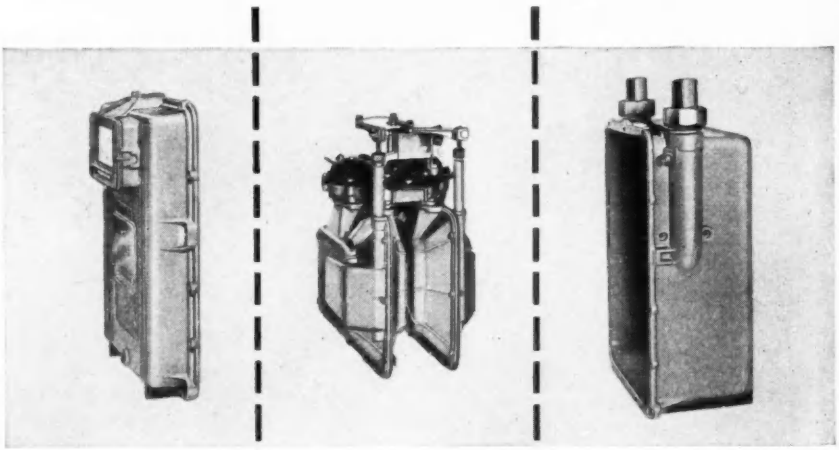

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Gas Journal

COKE OVENS • BY-PRODUCTS • CHEMICALS

Coal utilisation—have we been so slow?

THE report of the Wilson Committee on Coal Derivatives, published last week, makes it quite clear that the production of oil from coal is not commercially feasible, and advises that any technological research on the problems connected with it now in hand should be wound up as soon as possible. This applies, in the Committee's view, to the research now being undertaken on the Fischer-Tropsch process at the Warren Spring laboratories of D.S.I.R. They recommend that work there should be brought to a close within the next 18 months.

The Committee gives as its reason for this view that conditions during the war were such that these processes for the production of oil from coal were justifiable, but today, with the availability of oil products being so much greater and at a reasonable price, such processes are not economically viable.

A most important point about this report, which is the result of a Committee under the Chairmanship of Mr. A. H. Wilson, F.R.S., set up by Lord Mills, the then Minister of Power, in 1959 to enquire into other outlets for coal, is that until recently the coal industry has not had to worry very much in the post-war years about coal utilisation. Production has been its main problem and this it has pursued with all the energy at its command. It was not until 1958 that it had any difficulty in selling its product, though we had heard before this of doubts arising as to how to dispose of rising stocks of very small coal. Since 1958, with the contraction in sales of all grades becoming serious, steps had to be taken to find some means of using coal other than as a fuel.

If it could have been proved conclusively that oil from coal had been a commercial proposition, it seems to us that the vast sums required to complete the necessary technological research and to build plants large enough to play a significant part in the production of petrol and oil for home consumption, would indeed have been well spent. It might have gone some way towards making this country self-supporting in oil production and created new and certain markets for British coal.

The type of coal treatment which the Committee favour is some form of gasification, by which a fuel gas is made available to industry. It lays special stress on the development of the slagging gasifier, the research work on which should, it is thought, be made the responsibility of the Gas Council, and on the gasifier for small coal being developed by Ruhrgas A.G. in Germany. The responsibility for this work in Britain should be undertaken by the N.C.B.

Finally, the Committee do not think that apart from the chemical products obtained from coal carbonisation, there is much hope of increasing the annual total of 1.5 mill. tons of coal now used directly for chemical processes. The coal industry, in fact, appears to be very much where it was before.

A leading article in *The Times*, commenting on the report, complains that the gas industry has been slow in technical development compared with the electrical industry, and that the increased vigour it has shown in recent years has been spent in developing oil gasification and the importation of methane.

This may be a kind of back-handed compliment in that it suggests that the future of coal utilisation is bound up in our industry. If this is so, we accept the compliment, and we feel sure the industry accepts its responsibility as one of the chief processors of coal in the country. It must be remembered, however, that the gas industry is in business as a supplier of a refined fuel in competition with other fuel industries. To do this it must make every effort to produce gas and distribute it at the lowest possible price, which may not be possible by traditional coal carbonisation. The industry must be free to make gas by the most modern and up-to-date means that are available. It never has been an appendage of the coal industry.

To compare its development work in the post-war years with that of the electricity industry in this particular context is, we think, a little out of place.

The electricity industry has expanded over the last 30 or 40 years at a most striking and consistent rate. It supplies a refined fuel and a power and lighting service almost under the conditions of a monopoly, so that it probably has more resources to finance a very extensive research programme. It has taken full advantage of this research to build modern thermal power stations of ever increasing efficiency. The electricity industry can and does use lower rank slack coals for the firing of its high pressure, high efficiency boilers, which are considerably cheaper than the high rank gas-making coals. To compete on coal costs the gas industry must turn from carbonisation to gasification. The building

of the Lurgi plants in the Scottish and West Midlands Boards' areas is the first step in implementing this conception. Large quantities of gas should be made available for distribution over a wide area. A criticism may well be that work on this type of development was slow in starting, but this may be met by pointing out that only in the last ten years or so have coal prices forced a change of gas-making procedure and in that time, too, tonnage oxygen on a commercial scale has become available in this country.

Feedstocks and reforming plants

INITIAL operating difficulties, generally known to the engineer as 'teething troubles,' are a commonplace and are to be expected when any new piece of plant is being started up. But how much more serious, time-consuming and expensive can they be when a new plant employing some completely new process or principle is set to work for the first time!

A paper we are publishing this week, describing the early months of working of the Onia-Gegi oil and refinery gas reforming plants at Portsmouth and Southampton, by Mr. M. J. Birch, Senior Assistant Engineer at Hilsea works, Portsmouth, and Mr. H. J. Lockwood, Senior Assistant Engineer at Southampton, is most instructive in showing what kind of problems the contractors and the gas board engineers must be prepared to face when a new type of plant is started up.

We were most interested to see how great an effect the alteration of feedstock can have on a catalytic reforming plant. Three feedstocks were used at Southampton and two at Portsmouth; medium fuel oil and virgin naphtha were available at both, but Southampton alone had access to refinery gases from the Fawley refinery.

The ability to change over from one feedstock to another is one of the basic economic premises of the successful use of a reforming plant for oil distillation products, and without it, one of its chief advantages over coal carbonation is lost.

Clearly the heavier and more viscous the feedstock is, the less easy it is to handle, and the by-products from the plant in the form of tar and liquor become increasingly troublesome. Heavy, crude, slow running tars cause blockages in the most inconvenient places, and emulsions forming in washboxes and separators make the clear division of tar and liquor a problem.

We were, however, impressed by the way such obstructions, which were often serious enough to necessitate the closing down of the plant and demand a new approach to the problems created, were in fact surmounted by the authors. A change in operating conditions was found which modified the tar viscosity and certain reagents reduced emulsification.

Of the three feedstocks, virgin naphtha as received and without any previous processing, appears to have been the most satisfactory. It has a low sulphur content, it is easily handled and forms no tar. Since its carbon: hydrogen ratio is lower than that of medium fuel oil, less carbon is deposited during the reforming process, so that when used in the heating burners of the plant, a greater quantity must be burnt to provide preheat. The provision of secondary and tertiary burners to heat the catalyst beds was of interest, but here again care in the adjustment of flame length, when burning virgin naphtha, is important.

Logically the use of refinery gases should be simpler and more satisfactory than a light naphtha feedstock, but their high sulphur content is a great disadvantage, the effect of which is threefold: The poisoning of the catalyst, the adverse effect on the reforming efficiency, and the production of a most corrosive effluent. So serious did this last become that the Board was compelled to install a set of oxide purifiers to extract the hydrogen sulphide from the gas before reforming it.

Once this had been done, the difficulties with this feedstock were overcome and the fully designed output at the predetermined calorific value was easily achieved.

The use of an oil as heavy as medium fuel oil is of interest, though there seems to be a tendency to use the naphtha feedstock in increasing quantities. It is significant, however, that oils of this type and heavier might well become the kind of oil which the industry might have to gasify in the future. Gas and diesel oils are in demand, and the demand for fuel oils too is likely to rise. As long as virgin naphtha, under its many and various names, continues to be available, difficulties over the type of feedstock to be used are unlikely to be very serious. If the demand for this grade by the industry increases, however, it may no longer be a surplus product to the oil industry. Its suitability for the enrichment of c.w.g. and consequent advantage of being able to raise the output of any such plant designed for gas oil enrichment, goes a long way towards compensating for any but a most violent rise in price.

Refinery gases are clearly the most easily handled feedstock, provided they are first freed from H_2S. On this point the reforming plant at Romford was greatly helped when the North Thames Board decided to purify its refinery gas at source before it entered the high-pressure steel pipeline. It is interesting to speculate on the large quantities of corrosive effluents which might well have been formed with Romford's very large daily throughput, had this not been done.

We are grateful to the authors for bringing forward so many valuable points about the working of the Onia-Gegi plant and by so doing assisting those engineers likely to find themselves in charge of reforming plants at some future time. There are now a number of these plants in this country and on the Continent, and it seems that in the short term this number may well increase. In the long term the high pressure coal gasification plant worked in conjunction with underground storage for peak demand is likely to be the complete solution for long distance gas supplies. The discovery of natural gas on the Continent has brought such a system much nearer realisation, but in the meanwhile, with the cost of coal tending to rise and difficulties with coke sales, reforming plants capable of processing all types of oil feedstocks are likely to be in demand for the next few years.

From a paper to the London and Southern Junior Gas Association.

Catalytic oil gasification in the Southern Gas Board

By M. J. BIRCH,
SENIOR ASSISTANT ENGINEER, HILSEA WORKS, PORTSMOUTH,
and
H. T. LOCKWOOD,
SENIOR ASSISTANT ENGINEER, SOUTHAMPTON WORKS.

THE changing pattern of the gas industry since the end of the war, and the greater importance attached to oil and other refinery products, have been reflected in the production pattern of the Southern Gas Board. The geographical location of the Board also favoured expansion of oil gas rather than of coal gas.

The Board's first oil gas plant was installed at Portsmouth and began operation in 1956, and was the first self-steaming Onia-Gegi plant of its size to operate in this country. It has gasified both medium fuel oil and light distillate. The two units installed at the Southampton works began operation in 1958, and have used medium fuel oil, light distillate and refinery gas.[1] This refinery gas was originally reformed in carburetted water gas plants,[2] but in order to take full advantage of the maximum quantity available it was decided that the additional plants required would be Onia-Gegi plants.

Difficulties led to failure

The oil industry is subject to changing market conditions and so their products available to the gas industry follow this varying market. Some years ago medium fuel oil was the most attractive liquid fuel for gasmaking, but latterly light distillates have had certain advantages. Known by a variety of names or abbreviations, virgin naphtha is the product referred to in this paper. It is valuable, however, for a plant to be sufficiently versatile to gasify any of the feedstocks available, and those described in this paper have proved their reliability on varying feedstocks with little alteration to plant conditions.

Oil gasification is not a new process, but a fundamental requirement is the conversion of a carbon : hydrogen ratio of seven or eight in the oil to a final gas of carbon : hydrogen ratio of about three. Difficulties over this operation have caused the failure of many of the earlier processes. One method of decreasing this ratio is to add hydrogen by reacting with steam. This is actually carburetted water gas practice; the reactions, however, proceed very slowly under normal conditions, so the use of catalysts seemed the logical means of overcoming this difficulty.

The technical problem of producing a gas with the characteristics of normal town gas from oil and refinery gases has been the aim of several gas-producing plants in different parts of the world for some years. It is only recently, however, that attention has been given to the catalytic reforming of refinery gases in this country[1, 2].

The catalytic gasification of oil and the reforming of refinery gases is an extremely complex chemical process. Its primary function is to modify a hydrocarbon gaseous mixture from one of high calorific value to one of low calorific value suitable for domestic and industrial use. This process is carried out over a variety of catalysts in the presence of a hydrogen rich gas, *viz.*, steam, and may take various courses. The direction which the reaction will take is influenced by a number of factors.

Steam : feedstock ratio. The steam requirements for the gasification or reforming of the feedstock to carbon monoxide and hydrogen must in practice be higher than the theoretical in order to keep the deposition of carbon to a minimum. By increasing the process steam : feedstock ratio above this quantity a gas of higher calorific value is obtained. The excess steam might be expected to react with carbon and carbon monoxide to yield greater volumes of carbon dioxide, and so a gas of lower calorific value. It appears, however, that the effect of the increased volumes of steam is to protect the hydrocarbon gases and vapours as they pass through the catalyst beds by reducing the temperature of the reactants and increasing the space velocity. With this reduction in cracking there is less carbon deposited with less opportunity for high temperatures within the catalyst bed caused by carbon combustion during the heating phase.

Side reactions during cracking

Characteristics of feedstock. Feedstocks of higher carbon content and those requiring high temperatures for conversion will tend to produce greater quantities of free carbon. Previous authors[3] have shown that lighter feedstocks give an increased gas output at a higher gasification efficiency and the gas has a low specific gravity. This trend is substantiated by the present authors. The quality of the tar formed improves with the lighter feedstocks and the quantity is less, refinery gas producing a negligible amount. A feedstock, therefore, with a high paraffin content (virgin naphtha) yields the least quantity of tar and naphthalene, since the extent of undesirable side reactions during cracking is much less than with those containing a higher proportion of olefines, aromatics and naphthenes.

Operational temperatures and effect of thermal decom-

position. Operational temperatures throughout the plant and especially of the catalyst have a considerable effect on reforming and gasification efficiency, and the quality of tar produced.

Thermal decomposition of fuels

The thermal decomposition of the hydrocarbons commences as soon as the feedstock reaches temperatures over 650°C., and its degree depends largely on the hydrocarbon constituents of the feedstock. Hydrocarbon gases, oils and petrols (light spirits), contain aliphatic compounds (paraffins, olefines), aromatic compounds and naphthenes. Thermal decomposition of these constituents will take place in the catalytic plant by direct contact with the hot brickwork and in the free spaces of the plant vessels before reaching the catalyst. Because of their composition, however, refinery gases do not suffer thermal decomposition to the same extent as the heavier hydrocarbon liquid feedstocks, which undergo considerable thermal decomposition with the production of carbon in the vaporising chamber of the gas-making units. Milner and Taylor[4] have given a very full account of the theoretical aspects of thermal decomposition of hydrocarbon fuels in a recent paper.

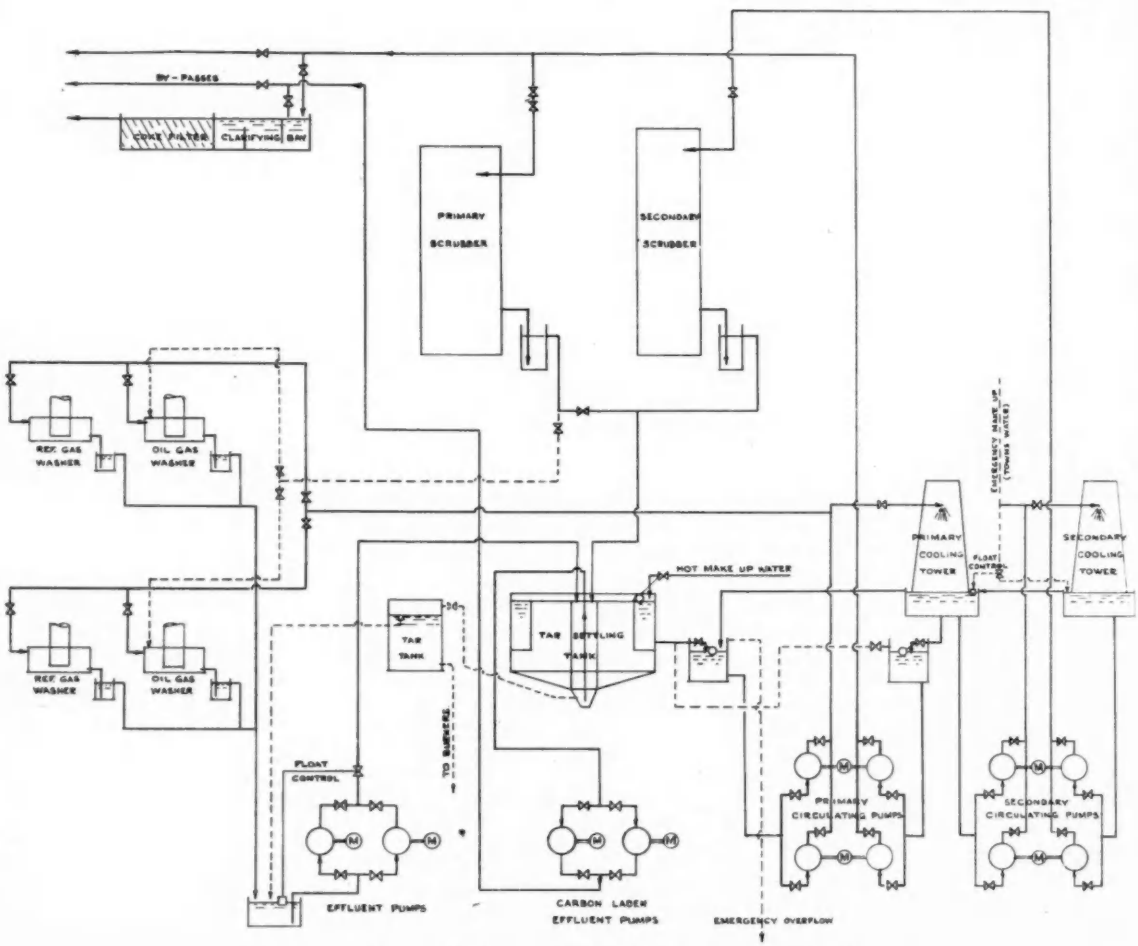
The extent to which thermal decomposition has taken place will impose limitations on the amount of catalytic gasification which can subsequently take place. The extent of thermal decomposition must therefore, be controlled to make certain of the maximum degree of catalytic gasification and so the greatest gas production. The degree of preheating of the feedstock in the vaporising chamber is, therefore, of some importance.

The presence of sulphur in feedstock. The conversion efficiency of nickel catalysts is seriously affected by sulphur compounds, especially when reforming refinery gases. The presence of hydrogen sulphide in the refinery gas feedstock of the order of 150 grains per 100 cu.ft. has been found to reduce the volumetric expansion considerably, and at 250 grains per 100 cu.ft. it has reached its minimum, *viz.*, some 40% of the theoretical. Any further increase in hydrogen sulphide has little effect.

Presence of sulphur compounds

There is no evidence on operational plants that the organic sulphur compounds contained in the heavier oils—900 to 1,000-second Redwood No. 1—affect conversion efficiency since sulphur free oil of this type is not available. Investigations, however, with test plants have shown that there is an increase in carbon deposition from feedstocks of higher sulphur content.

When gasifying almost sulphur-free virgin naphtha—.04 to .1% by weight—at Portsmouth, but still using medium fuel oil (900 to 1,000 second's viscosity and containing 3.5 to 4% organic sulphur) for the burner fuel, the



Diagrammatic layout of effluent and circulating water system, Southampton.

conversion efficiency was below that anticipated. With a change to virgin naphtha for burner fuel, conversion efficiency increased by 25 to 30%. The presence, therefore, of sulphur compounds in the burner fuel, as well as in the feedstock, appears to reduce the conversion efficiency. This has been further substantiated by injecting sulphur dioxide into the heating phase when reforming refinery gas. The experiments showed that a concentration of 200 grains of sulphur dioxide per 100 cu.ft. in the heating gases considerably reduced the volumetric expansion.

It has been suggested that the large quantities of carbon deposited when gasifying the heavier oils does in fact afford some protection from the poisoning of the nickel catalysts by sulphur. Analyses of the sulphur compounds in the product gas from the heavier oils show a high percentage of carbon disulphide and carbon oxysulphide, which suggests the combination of sulphur and carbon.

The effect of organic sulphur compounds in the refinery gas feedstock on nickel catalysts has still to be investigated. An organic sulphur content of from 5 to 8 grains per 100 cu.ft. did not affect conversion efficiency in the test plant, and so far has not affected the operational plants at Southampton.

Rate of reaction with steam

Rate of reaction with steam. Liquid feedstocks accompanied by steam in the presence of a catalyst at an elevated temperature, break down into numerous products. Hydrocarbons formed by earlier thermal decomposition in the plant will be further broken down and the final gas made will be a mixture of products, the formation of which has been very complex. The steam appears to react with the cracked hydrocarbon fragments of free radicals and it is the reactions between steam and these free radicals which is catalysed.[4] The rate of combination of the free radicals and their catalysed reaction with steam will therefore have a considerable influence on the characteristics of the gas produced.

The mechanism of the reaction between steam and gaseous feedstocks is complex, but appears to be simpler than with liquid feedstocks, though the reactions are similar in both cases. The reforming of gaseous feedstocks can be taken to completion, which suggests an easier and less complex process. Typical reactions with steam are: Liquid feedstocks,

$$C_xH_y + xH_2O = xCO + y/2+xH_2.$$

Gaseous feedstocks,

1. $CH_4 + H_2O = CO + 3\,H_2$
2. $CH_4 + 2H_2O = CO_2 + 4\,H_2$
3. $C_3H_8 + 3H_2O = 3CO + 7\,H_2.$

In addition, the water gas reactions take place when carbon is present.

The quantity of carbon being deposited. The quantity of free carbon released from the feedstock depends upon the degree of thermal decomposition taking place, the quantity of process steam used and the activity of the catalyst. The greater the amount of carbon left behind in the plant vessels and appearing in the tar, the lower the output from the plant. But the production of carbon, when gasifying the heavy liquid feedstocks, is unavoidable if the maximum amount is required at the lowest calorific value. While reforming refinery gases, carbon deposition appears to vary with the carbon content of the gas. Although a high calorific value gas, rich in hydrocarbons, deposits the greater amount of carbon, it has been found that carbon is not easily deposited with refinery gases, and that a variation in operational temperatures and steam: gas ratios does not have the same detrimental effect as on liquid feedstocks.

Space velocity. This is usually measured as weight or volume of reactants per hour per unit weight or volume of catalyst in the reactor. A decrease in space velocity increases the contact time and so causes greater conversion and a lower calorific value, but not necessarily with greater efficiency since carbon deposition may be increased.

The degree of oxidation and rate of reduction of the catalyst. When nickel catalysts are used, oxidation and reduction of the nickel takes place during the heating and gas-making phases respectively, so as the quantity of nickel oxidised increases more oxygen will be removed in the later initial stages of the gas-making phase. This quantity of oxygen will later react with one or more of the constituents of the feedstock or product gas, *i.e.*, hydrogen, carbon monoxide, carbon and sulphur compounds. Typical reactions would be:

1. With hydrogen:
   $NiO + H_2 = Ni + H_2O.$
2. With carbon monoxide:
   $NiO + CO = Ni + CO_2.$
3. With carbon:
   $2\,NiO + C = 2Ni + CO_2.$
4. With hydrogen sulphide:
   $3\,NiO + H_2S = 3Ni + SO_2 + H_2O.$

The effect of these reactions is that reaction 1 represents a loss in hydrogen and therefore output, reaction 2 represents a loss of potential heat with an increase in inerts of high specific gravity, reaction 3 causes an increase in inerts and a loss of carbon, and reaction 4 produces a gas with objectionable acid constituents.

Experiments carried out at Portsmouth showed that it was possible to reduce the tendency for reactions 2 and 3 to take place by limiting the amount of excess air introduced during the heating phase. Care must be taken, however, when limiting the excess air requirements, to ensure that no carbon is left behind unburnt on the catalyst at the end of the heating period.

Activity of catalyst decreases

Activity of the catalyst. Catalysts impregnated with metallic nickel have the property of greatly accelerating the interaction of hydrocarbons and steam, as is indicated by their ability to promote the reaction between steam and methane at relatively low temperatures. The nickel catalyst may therefore be used for the complete reforming of refinery gases, containing a large proportion of methane, to hydrogen and carbon monoxide.

The activity of the catalyst increases with rising temperature but a point is eventually reached when the formation of carbon reduces its efficiency and the activity of the catalyst diminishes with any further increase in temperature.

Freshly prepared catalyst has a very high activity because of its high surface area, but the activity falls off in time due to contamination and the reduction in surface area caused by sintering under the effect of heat and steam.

The gasification process is cyclic, involving heat and gas-making phases, separated by suitable purge periods.

The heating (or blow) phase is used, initially, to bring the plant from cold to operating temperatures, and thereafter to restore the heat content of the plant following each gas-making phase. The heat required for this is provided by the plant burners; the primary burner in the combustion chamber, with its associated primary air, is the main burner and the secondary burner with its secondary air, is used to balance the temperature of the catalyst bed. However, by using this secondary burner to its maximum effective spraying capacity, greater overall

efficiency is obtained. The pre-heater must provide sufficient heat for vaporisation of the feed stock, and this the primary burner does.

The heat requirements for medium fuel oil are obviously greater than when light distillate is used, so there is a greater overall efficiency when gasifying the latter feedstock.

The Southampton plant units have two catalyst beds, and a tertiary burner with its air supply is provided for maintaining the temperature of the lower bed. Another feature of the Southampton plant is the provision of an additional air supply at the back end of the combustion chamber to assist in the combustion of carbon and to spread the heat in the pre-heater when the heavier feedstocks are gasified. The burning of the carbon deposited during the make phase provides a considerable amount of heat for the process. Air in excess of the theoretical quantity is supplied to ensure that all carbon deposits are burnt off and that a 3% oxygen content in the waste gases is maintained.

Purge periods necessary in cycle

After a short steam purge the gas-making phase commences. When gasifying liquid feedstocks steam passes through the pre-heater, and meeting the oil vapours sprayed in at the top of this vessel, carries them through the catalyst bed. At Southampton when reforming refinery gases the whole of the process gas, together with about 60% of the process steam, enters the pre-heater above the chequers. The remaining 40% of the process steam enters below the chequers as with liquid feedstocks. Some thermal cracking occurs before the catalyst bed, depending on the temperatures inside the plant, but the main reactions occur in the catalyst bed where the steam and hydrocarbons react to form lower hydrocarbons, carbon monoxide, carbon dioxide and hydrogen. Carbon is deposited and the water gas reaction also takes place. The amount of steam used depends on the steam: oil/gas ratio required. As this phase proceeds, the temperature of the catalyst bed falls, and its activity is reduced, causing an increase in the calorific value of the gas made. A return to the heating phase is then necessary to restore the plant temperatures. At the end of the gas-making phase the catalyst is reduced.

Purge periods follow each phase of the cyclic operation, to clear the set of blow gases before gasmaking and to sweep away residual make gases before the heating phase. The time involved should be kept as short as possible consistent with safe operation.

Both plants are equipped to operate on four-, six-, eight- or 12-minute cycles. At Southampton all the work has been on a six-minute cycle; at Portsmouth using medium fuel oil, the plant operated on a six-minute cycle, but this was changed to four minutes for light distillate. As with carburetted water gas practice, the best results are obtained on a short cycle, since temperature conditions are usually more stable. The limiting factor is the length of time required for purge periods and valve movements.

The choice of cycle and how it should be divided between the various phases depends on a variety of factors, among which are:

1. Sufficient time must be available during the heating phase

(a) for burning off the carbon deposited during the gas-making phase without rapid combustion causing an undue rise in catalyst temperature,

(b) for allowing the restoration of plant temperatures and the heat capacity of the plant,

(c) for keeping the air supply within the margin of the air blower, and

(d) for allowing synchronisation when two units are working in parallel.

2. Sufficient time must be available during the gas-making phase, (a) to ensure the correct contact time for efficient catalytic reforming of the feedstock and to obtain the maximum output of reformed gas, (b) to keep the temperature drop to a minimum throughout the period, and (c) to allow adequate purging of the vessels.

The plants can be operated through a wide range of output, and quality control may be adjusted to suit requirements. Each plant is designed for a given output and corresponding space velocity, so any reduction in output should be achieved either by shortening the oiling period or by intermittent working, and not by reducing the oiling rate.

Both plants have gasified various feedstocks. The Portsmouth plant changed from medium fuel oil to virgin naphtha and only alterations in cycle time were made. At Southampton where gaseous feedstocks have been used, a more active nickel-impregnated catalyst on a sillimanite base is necessary for reforming refinery gases, while a nickel-impregnated catalyst on a magnesite base gives the best results for the medium fuel oil.

The process must be controlled so that (a) the gas made is of the required calorific value and has the correct composition, with a minimum of inerts, sulphur compounds, naphthalene or other undesirable constituents, (b) any tar produced is sufficiently mobile to be easily handled, (c) the operation is continuous for the maximum length of time, so that carbonaceous deposits in the plant are kept to a minimum, (d) carbon deposition in the catalyst bed is under control, (e) catalyst temperatures are kept at a safe level, (f) smoke emission is avoided, (g) gas-making efficiency is at a maximum, and (h) there are the correct combustion conditions during the heating phase.

Each of these items is inter-related and good operation achieves their optimum combination. Any alteration should be small and sufficient time allowed to observe its effect before making any further change.

The primary factors influencing plant operation are calorific value and the volume of gas made.

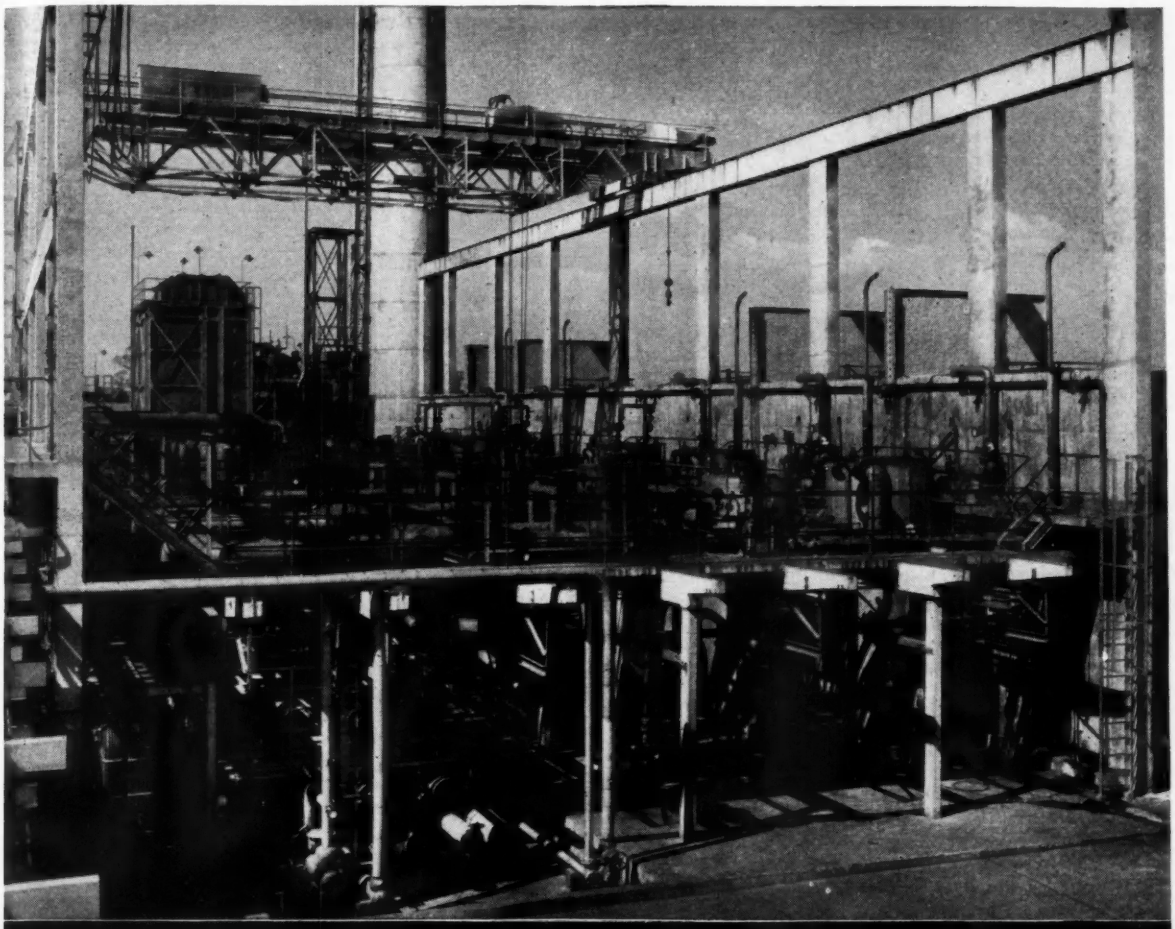
Control of carbon deposits

For a given calorific value there are definite rates of oil and steam flow. On the self-steaming plant, calorific value is mainly controlled by an alteration of the catalyst temperature, but change in the steam: feedstock ratio is also a very effective means of quality control. Thus a reduction in catalyst temperatures increases the calorific value of the reformed gas and an increase of the steam: feedstock ratio will also increase the calorific value. The calorific value may be reduced by reversal of these procedures.

This is brought about by an increase in temperature causing additional reactions with steam and *vice versa*, while an increase in steam: oil ratio may cause a reduction in the temperature either of reactants passing through the catalyst or of the catalyst bed itself, by preventing carbon being deposited.

Tar quality depends on the feedstocks used and the plant operating temperatures. High catalyst temperatures usually produce a thick tar. A small increase of the steam: oil ratio helps to remedy this but it may raise the calorific value of the gas above that required. Low temperatures produce a thin tar, and if the calorific value of the make gas is correct this is the best means of producing a manageable tar. The maximum use of the secondary burner, and so reducing the amount of thermal cracking, solved this heavy tar problem at Portsmouth. Carbon build-up in the catalyst is another cause of viscous tar.

Carbon deposits need not cause any problems if the correct amount of excess air is maintained. These deposits are mainly associated with medium fuel oil working and



High Pressure Oil Gasification at Shell Haven

The Shell Oil Gasification Plant erected by Woodall-Duckham Construction Company Ltd. at Shell Haven for the Shell Chemical Company Ltd. The plant produces synthesis gas used for fertiliser production.

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CONSTRUCTION COMPANY LTD
Woodall-Duckham House · 63-77 Brompton Road · London · S.W.3
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build up on the pre-heater chequers and in the catalyst bed. Unsatisfactory oil spraying and low air or steam rates lead to a build up in the pre-heater. If the condition becomes severe the carbon will flare during the heating phase and cause considerable smoke emission.

Since hydrocarbons break down with a deposition of carbon, depending on the feedstock, type of catalyst and operating conditions, those feedstocks of high carbon: hydrogen ratio favour carbon deposition. Even flow distribution of oil vapours, gas and steam through the catalyst is essential, while poor distribution causes different temperature levels in parts of the bed, and varying quantities of carbon to be deposited. As the temperature of any part increases, more carbon is deposited until there is insufficient air passing through that portion of the bed to burn it off, or the catalyst is damaged by high local temperatures. Under such conditions the use of feedstocks of high carbon: hydrogen ratio, and the back pressure resulting from carbon and ash deposits will lead to insufficient air being available to give a safe excess. It is then necessary to shut down and screen the catalyst, rejecting all the dust and smaller particles.

High temperatures increase the activity of the catalyst and favour the water gas reaction between steam and deposited carbon. They also promote the thermal breakdown of hydrocarbons, which can reduce efficiency by excessive carbon deposition.

Protection of the catalyst

The catalyst is robust, with a considerable margin of refractoriness, but it is important to guard against conditions which may cause its premature failure. Excessively high temperatures during heating must be avoided, and the excessive use of the secondary burner causing flame impingement, can cause a breakdown of the catalyst.

For any feedstock the operating temperature is governed by (a) the catalyst activity, (b) the quantity of catalyst and so the space velocity of the gas, (c) the steam to oil ratio, and (d) the calorific value required.

If these conditions are pre-determined then the catalyst temperature is constant, and any adjustment will upset the fixed conditions. Operating temperatures should be as high as possible, but limited by:

1. The excessive use of steam to protect the hydrocarbon from thermal cracking and to maintain the calorific value of the make gas.
2. The heat losses in the gas at the inlet to the washbox caused by high operating temperatures.
3. The mechanical strength of the catalyst.

During the heating phase, excess air is used and in the early stages the combustion of carbon causes high temperatures on the catalyst surface, but in the self-steaming units this excess air is diluted by the heating gases and only about 8% of oxygen enters the catalyst bed. This minimises the overheating dangers.

The catalyst is reduced during the gas-making phase and oxidised during the heating phase. With impregnated catalysts over-oxidation is thought to cause local hot spots; the catalyst becomes sulphided and sulphated during the make and is then more readily oxidised during the following heating period with further rise of temperature. If this continues, then catalyst fusion will occur.

A highly oxidised catalyst causes a high carbon dioxide content of the make gas, especially in the early part of the make period, but this can be obviated by reducing the excess air, but it must be stressed that this was only attempted when gasifying light distillate.

Smoke emission is caused by (a) an incomplete purge of the make gas, (b) insufficient excess air to burn off the carbon deposit, (c) the incomplete combustion of the burner oil, (d) poor atomisation of the burners, (e) a build-up of carbon in the pre-heater, caused by poor spraying of the oil for gas making, and (f) low temperatures in the pre-heater or outlet catalyst, causing incomplete combustion during the early part of the heating phase.

There should be no difficulty in controlling these items. Particles of carbon are carried forward with the waste gases and can be a considerable nuisance. Since most of these are precipitated in the waste heat boiler, regular cleaning of the dust pockets is essential to minimise dust emission. Some of these particles, when crushed, are found to contain an oily portion, which may possibly be due to incomplete vaporisation. When gasifying virgin naphtha or refinery gas, however, there is no problem of smoke or dust emission.

The height of the chimneys—110 ft. (1956) at Portsmouth and 210 ft. (1958) at Southampton—makes them noticeable landmarks. The height is in accordance with the requirements of the Chief Alkali Inspector, and is reckoned on the gasification of medium fuel oil from a plant twice the capacity of the present installations, *i.e.*, a 100% plant increase would be possible without any further increase in chimney height or capacity.

The difficulties encountered in the early stages of plant operation when gasifying medium fuel oil are well known. Since the Hilsea plant at Portsmouth was one of the early plants (1956), it is possible that some troubles were peculiar to this unit alone.

The formation of a very viscous tar which defied manual handling, was the first and most prolonged difficulty. This tar was completely immobile and flume blockage caused more plant shutdowns than any other trouble. Tar trouble also appeared in the washbox, and excessive back pressure here caused more shutdowns.

The detarrer was subjected to considerable overloading. The design loading was 33 kVA. but the instantaneous load reached 44 kVA. and condensation occurring in the insulator chamber caused nine insulators to fail.

Problem of tar viscosity

The slip of tar fog from the detarrer bonded together with the naphthalene to form a sludge, causing excessive back pressure on the scrubber and naphthalene washer. Since this sludge found its way into the separator system, the cooling tower also gave very poor results, which were due to blocked troughs. The separator system was originally in two sections, one receiving the washbox and detarrer flume and the other the scrubber flume. The viscosity of the tar in the separators presented problems of pumping to the sales tank, but since laboratory investigations showed that naphthalene mixed with tar reduced its viscosity very considerably, the two separator sections were then merged and easier tar handling followed.

The problem of the thick tar was solved by working with lower pre-heater temperatures to minimise the thermal cracking, and by greater use of the secondary burner to obtain the correct catalyst temperatures. Surface agent additives were used in the washbox to assist in breaking down emulsions.

The detarrer problem was investigated and a new type of graphite lined insulator was fitted, together with an aluminium shield which acted as a reflector. The clean gas supply was re-designed, a ring main was installed with a pre-heater, and clean gas entered the chamber at four points of the circumference.

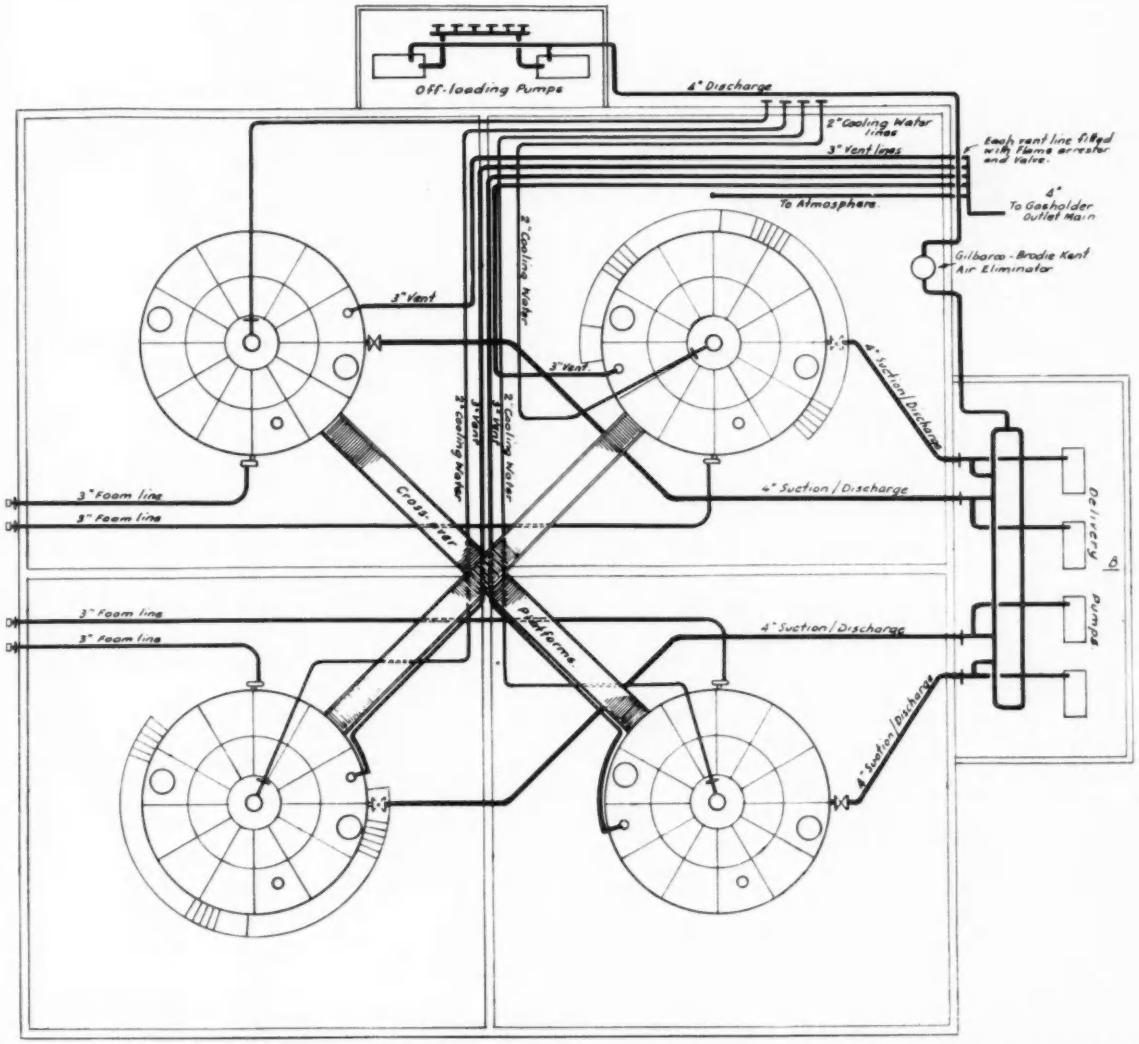
To overcome the sludge problem the angle of the outlet main from the scrubber was altered to give a steeply inclined fall, a wood lining was fitted, and banks of sprays to operate cyclically were installed. This partly overcame the problem, but it tended to cause flop-over into the oil

washer, so a drain leg was fitted with a seal pot. These modifications solved the problem of sludge and no further back pressure difficulties were encountered.

The deposition of carbon in the catalyst vessel and the resulting build-up of back pressure implied that the time between catalyst removal and screening could affect the operating time available. Air rates were carefully arranged to yield 3% oxygen in the stack gases over the complete heating phase, and so ensured that there was always sufficient excess air to burn off carbon deposits.

The introduction of virgin naphtha was in two stages. First as a gas-making feedstock only, with medium fuel oil as a burner fuel. No cycle alterations were made and there were no operational troubles. However, the *p*H values of the circulating system began to fall and a dosage of the system was begun. The dosing agent chosen was ammoniacal liquor, a supply of which was readily available. This emphasised that carbon does protect the catalyst from sulphur attack. Operating under these conditions the SO_2 content of the early phase of the make gas was 446 grains per 100 cu.ft. and as mentioned earlier, conversion efficiencies were lower than anticipated. When virgin naphtha was later used as a burner fuel these *p*H and SO_2 problems disappeared and the plant operated with very little trouble.

The proper use of the secondary burner is very important and has significant influence on plant operation. When gasifying virgin naphtha less heat is required in the pre-heater, and therefore, if this secondary burner could be used at a high rate, the overall quantity of burner oil per cycle could be reduced. Great care is needed, however, to avoid flame impingement on the catalyst since this can cause fusion. In the summer of 1959 an opportunity to shut down for inspection and maintenance presented itself; it was evident that high local temperatures had occurred and part of the catalyst bed had suffered from flame impingement. Consequently a 3-in. deep layer of broken firebrick was laid on the top of the catalyst as a protective layer and long thermocouples were fitted in order to find the temperature distribution across the catalyst mass.



Plan view of the virgin naphtha storage tanks, showing the delivery and suction pipelines, the vents to the outlet main of the gasholder, and the foam line connections to be used in case of fire. The whole installation is surrounded by an unclimbable safety fence.

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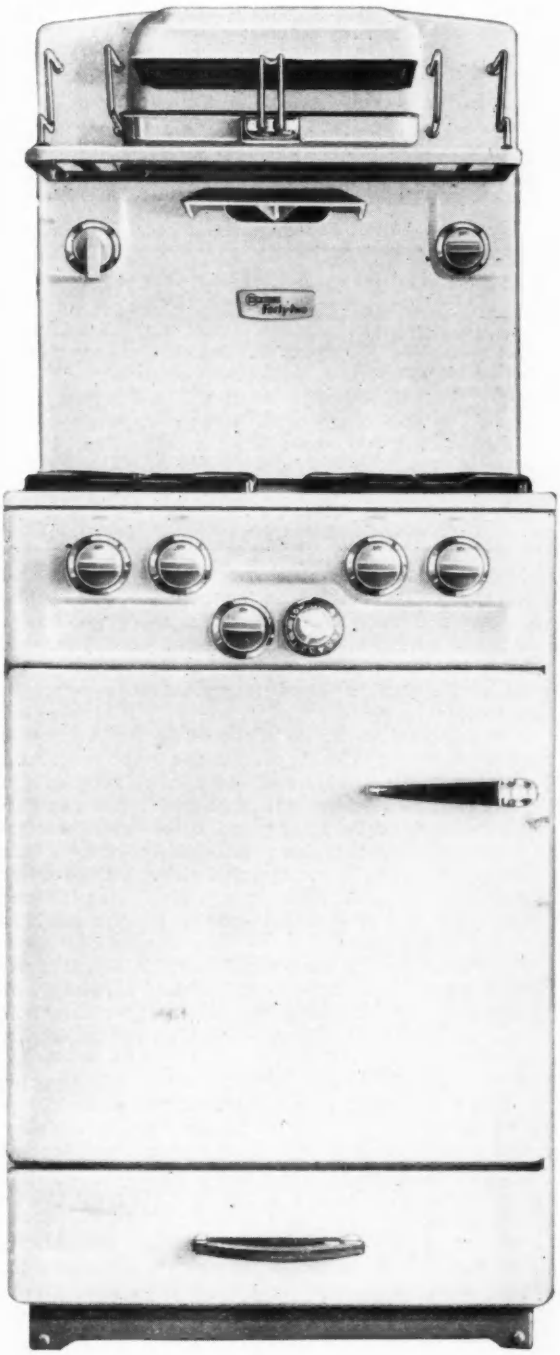

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The absence of Conradson carbon in virgin naphtha meant that extra heat had to be provided during the heating phase. Using a six-minute cycle the temperatures tended to fluctuate considerably, up to as much as 50° per cycle in each recording position, so it was decided to try a four-minute cycle. This proved to be very advantageous, the temperatures being maintained with almost straight line regularity.

The proper proportioning of air when virgin naphtha is used, is vital since short flame conditions can easily obtain with disastrous effects to brickwork. The primary air must be adjusted to give a long flame, and the steam pressure on the burner regulated to help to spread the flame.

The decision to operate the turbines against a back pressure of 20 p.s.i., using the exhaust steam for the carbonising plant was theoretically correct, but the steam demands of the plant frequently caused fluctuations of the works steam pressure and interfered with the turbine speeds, and consequently blast pressure. This may not appear to be connected with operational experiences, but it does show that the Onia Gegi plant is not self-supporting in steam production, and that other works functions must be taken into account when considering its operation. The works steam balance must, therefore, be planned accordingly to meet its requirements.

The first unit at Southampton, the No. 2 plant, was gasmaking with refinery gas in November, 1958, but it was immediately apparent that there was very little catalytic reforming taking place on the top catalyst bed of magnesite and that the bottom bed of sillimanite was not carrying the reforming to completion. On the fifth day a high acidity in the circulating water system was discovered, when *p*H values of 3 to 3.5 were measured in the washbox and primary scrubber waters respectively.

Both the poor catalytic reforming and acidity of water were attributed to the high H_2S content of the refinery gas—between 1 and 2%. It was believed that during the initial stages of the make period, the oxidised catalyst was being reduced by the H_2S; the SO_2 so produced, together with the unaffected H_2S, was being absorbed by the washbox and scrubber waters to form H_2SO_3, H_2SO_4 and solutions of H_2S, and so accounting for the highly acidic solutions.

Presence of high acidity

An inspection of plant vessels, etc., showed considerable graphiting on cast iron pump impellers and casings; mild steel pipe work had been corroded away and even stainless steel parts had been pitted.

The circulating water was, therefore, changed and gasmaking re-commenced with soda-ash dosing of the circulating water. Highly acidic conditions soon appeared, however, and the plant was shut down after only 18 hours of work.

Since the plant was intended to be operated on either medium fuel oil or refinery gas, a trial with medium fuel oil began in mid-December, 1958, but difficulties with carbon build-up on the steam pre-heater, chequers and viscous tar necessitated shutting it down again. These troubles, attributed to the design of make-oil spray, seemed to be surmountable.

In the meantime further consideration had been given to the acidity problem with refinery gas reforming and this was overcome by diverting the highly acidic water from the primary scrubber and washboxes to waste. The degree of acidity was thus kept reasonable and undivided attention could be given to the gas production problems.

Gas production from refinery gas was continued in order to explore all the problems giving rise to acidity, and by January, 1959, numerous cyclic determinations of H_2S and SO_2 showed that large volumes of SO_2 were being produced during the first minute of the make-period—900 grains per 100 cu.ft.—the H_2S remaining at about 100 grains per 100 cu.ft.

In order to avoid this large quantity of SO_2 entering the water system the stack closing cam was advanced so that the period of refinery gas admission in relation to 'stack closed' was delayed by 8%, as shown in Table I.

TABLE I.

	Original Cycle % of Cycle	New Cycle % of Cycle
Primary burner open	97	94
Stack open	0/100	0/100
Primary burner shut	38	36
Steam open	38	36
Stack shut	40	44 (+4%)
Gas on	42	38 (−4%)
Gas off	89	84
Steam off	97	94
	Gas on 2% after stack shuts.	Gas on 6% before stack shuts.

These alterations increased the amount of SO_2 to the stack and reduced the SO_2 in the make gas, and enabled the closed circulation system to become possible again.

With the new cycle a further series of gas analyses and sulphur determinations were made, but more attention was given to the factors which might influence reforming efficiency.

A study was made of the effects of (a) the variation in steam : refinery gas ratio, (b) the duration of the gas making period, (c) the temperature of the catalyst beds, (d) the influence of the rates on carbon black formation, (e) the sulphur content of the stack gases, (f) the variation in the rate of admission of refinery gas, (g) the effect on the catalyst of injection of medium fuel oil for a short period immediately before the stack closes, and (h) the effect of a cold blow period of ½-minute duration between the blow and gas-making periods.

From these experiments a good deal of information was gained, but no improvements in reforming efficiency were obtained. It was, therefore, decided that purifiers were needed to remove the hydrogen sulphide from the refinery gas.

By the end of February, 1959, the second Southampton unit (the No. 1 plant) was put to work. Medium fuel oil was gasified with the aim of working both units alternatively on oil during the refinery maintenance period, when the available quantities of refinery gas were only sufficient for thermal reforming in the water gas plant and for cold enrichment.

This period gave disappointing results. The bottom catalyst beds, being of sillimanite, appeared to be limiting the plants' performance, not only because of their greater activity and higher tar producing characteristics but also because their maximum permissible working temperature of 850°C. precluded adequate top bed temperatures. The decision was made to change the bottom bed on one unit (No. 1 plant) to that of magnesite with 3-in. extra depth, so that unit had two similar beds. After 12 days of continuous work this unit was shut down due to heavy tar and carbon choking the wash box. When the second unit was prepared for gasmaking, while the No. 1 plant wash box was being cleared, it was found that the top catalyst bed had almost completely disintegrated to a white powder, so this unit was rendered unworkable and its catalyst was renewed.

While the top catalyst was being renewed, gasmaking was

restarted on the other unit, but serious tar difficulties continued in spite of the fitting of water sprays on the inlet of the wash box, and several shut downs followed for the removal of tar from the washer and the flume.

Further working on medium fuel oil was temporarily abandoned. With the increasing volume of refinery gas becoming available once more, it was decided to return to the reforming of refinery gases. In May, 1959, both units were put to work and were operated at reduced outputs until a purification plant for removal of H_2S from the refinery gases was installed.

Refinery gas continued to be used until a supply of virgin naphtha became available in October, 1959, when each unit in turn was converted for the gasification of this feedstock.

During the long period of reforming refinery gas from May to October, the problem of acid formation, although partly arrested, still persisted in the gas scrubbers and the effluent flume. This was due to the pick-up of acid gases by the wash water in the scrubbers where no direct soda ash dosing could be effected. The plant had to be shut down on several occasions for the renewal of pipework handling this acid water and also for patching holes which appeared at the base of secondary scrubbers.

Increase in catalyst activity

The short period of virgin naphtha reforming on both units was satisfactory, although the rated output was not reached when medium fuel oil, which contained sulphur and so affected catalyst activity, was used as a burner fuel. Operation on this feedstock was very easy and no difficulties were experienced with tar or carbon.

During November 1959 both units were shut down for inspection and repairs. The opportunity was taken to change the catalyst on one unit (No. 1 plant) from two beds of magnesite to two of sillimanite, ready for working on purified refinery gas. The top catalyst bed had also a protective layer of broken fire brick—1-in. to 1½-in. cubes—added for protection from flame impingement.

In mid-December 1959 H_2S-free refinery gas was used for both burner fuel and feedstock in the unit with the top catalyst bed of magnesite and a bottom bed of sillimanite (No. 2 plant). The effect of the H_2S-free gas was immediately apparent within minutes of its admission; the H_2S content of the reformed refinery gas had fallen from 90 to 20 grains per 100 cu.ft. and the effluent scrubber water acidity began to fall within an hour, which indicated that there was a reduction in the amount of SO_2 produced. After 24 hours' working the catalyst activity had increased to the extent that the volume of reformed gas was approximately 40% higher than previously with the same volume of feedstock; its calorific value was reduced by 100 B.t.u. per cu.ft. and so the volume of enriched reformed gas was considerably greater. The increase in catalyst activity allowed the temperatures to be reduced, and there was a noticeable decrease in naphthalene formation.

The second unit, with its two beds of sillimanite (No. 1 plant) was put to work under conditions whereby 10 mill. cu.ft. per day of enriched reformed gas could be expected. This volume was immediately attained as anticipated with the H_2S-free refinery gas.

The catalysts used at Portsmouth and Southampton are magnesite or sillimanite, of which the following is a brief description.

Sillimanite:

Al_2SiO_5	Nickel	3%
	Outside layer	3% Ni.
	Inside layer	2.7% Ni.
	Calcium	Trace.

Surface area: 11,280 sq.in. per cu.ft.

Hydration test: Loss in weight 0.03%.

Magnesite:

Nickel	3.6%
Outside layer	10.0% Ni.
Inside layer	1.0% Ni.
Calcium	Trace.

Surface area: 6,000 sq.in. per cu.ft.

Hydration test: Loss in weight 0.15%.

During the hydration test, the catalyst is fired for one hour at 1,000°C. and is then autoclaved at 10 p.s.i. for two hours. It is placed well above the water level so that it is in contact with steam alone. Baffles are used to avoid splashing. Similarly, the catalyst container must be proof against liquid condensing on it. After the autoclave test the material is dried at 95°C. for one hour. Loss on ignition is then determined at 1,000°C.

Sillimanite catalyst is supplied in the form of small solid cylinders; dimensions vary from lump to lump. Magnesite catalyst is in irregular shaped pieces, dimensions again varying.

The catalyst is a calcined mixture, graded to size, and the catalytic effect of magnesite is concentrated on its outer surface. Control of impurities in materials used for its manufacture is most important. Small amounts of undesirable materials may be released from a catalyst and deposited elsewhere in the plant. Some metal oxides may be reduced at high temperatures and carried as a vapour to a lower temperature zone; here the reaction may reverse and deposit a powdered metal oxide. A few pounds of this type of material could plug a packed reactor vessel.

The catalyst must possess structural stability at high temperatures, which may reach 1,000°C. and local hot spots may rise to 1,200°C. Thus the catalyst support must be highly refractory, capable of standing this heat without impairing its chemical nature. Similarly, impurities in the oil used for feedstock can cause low fusion temperatures.

Heat effects on catalysts

Catalysts are subject to thermal shock. High compression stresses occur on heating and cooling. If spalling occurs fragments can be carried forward in the gas stream and be lost, or remain in the vessel and add to the back pressure. In selecting supports, heat capacity is an important consideration. Although low density supports may be more practical, the accompanying low heat storage properties may be detrimental. Generally, supports of high porosity and low bulk density require a greater weight of nickel but a lower total weight of catalyst for a specific conversion duty.[5]

Tests in the United States have proved that when 5% nickel was impregnated on various supports, magnesium oxide gave exceptional and uniformly high activity. Alumina gave high ratings only in certain cases.[6]

Metallic nickel promotes the reforming of methane at low temperatures and promotes the steam hydrocarbon reactions. It has been suggested that an increase of nickel content reduces the amount of carbon deposition, but at this stage the authors cannot contribute to this statement.

If overheated (above 1,100°C.) a nickel catalyst diminishes in activity, therefore burning off processes need great care. A safety margin of refractoriness and good mechanical strength, together with resistance to thermal shock are essential properties. The shape should give a high ratio of surface area to volume; this is to provide a large surface for reaction and easy heat transfer.

The magnesium oxides prepared by heating $MgCO_3.3H_2O$ and $(NH_4)_2CO_3$. $MgCO_3.4H_2O$ under identical conditions at 810°C. for one hour are hydrated to 70% and 86% respectively when left in water for six hours, while the

material prepared from the nitrate was only 43% hydrated. The higher the temperature at which the oxide is formed the less its reactivity to water and mineral acids. Variation in activity may be due to varying particle size, governed by temperature formation.

It follows that if magnesium oxide is heated strongly after it is formed, its affinity for water is affected. It may also explain why some magnesium-based catalysts have greater resistance to steam than others. Those formed at higher temperatures, or from sources giving a less reactive oxide, probably provide more stable catalyst supports.

The plants described in this paper have not been in operation for a long period, but from experience gained it is anticipated that the life of the catalysts will exceed that normally expected which, according to some authors, is two or three years.

The use of virgin naphtha

The catalyst from the Portsmouth plant was examined after three years working life, in which time the plant had made 3½ mill. therms, and only 15% was rejected on a size basis. The remainder was re-nickelled and made up to quantity with new catalyst. It must be remembered that this was the original catalyst and the plant one of the earliest in the country. Future installations should exceed this performance, especially with the use of protective layers.

Using correct air rates there need not be any uneven temperature distribution in the catalyst bed. Carbon build-up, if uncontrolled, can cause irregular and very high temperatures, with local hot spots. The pressure loss through the bed is caused by carbon and ash build up, and also by deposits from spalling. The degree of tolerable back pressure depends on the margin of the air blowers.

The effects of sulphur have already been mentioned. Catalyst life and activity vary with the amount and type of sulphur present. The performance of nickel catalysts in the presence of sulphur can be improved by careful adjustment of feed rates, steam rates and temperature. Hydrogen sulphide causes low catalyst activity and can cause carbon deposition, but no permanent harm need be caused if regenerations are carefully controlled, avoiding excessive temperatures. Sulphur poisoning shows itself by (a) increased carbon deposition, and (b) lower conversion efficiencies.

The reactions occurring in the presence of sulphur are generally shown as:

$$Ni + S = NiS.$$

$$2\,NiS + 3O_2 = 2NiO + 2SO_2.$$

However, the Southern Gas Board's central laboratory has found that more complex reactions may occur.

$$2NiO + 2S = 2NiS + O_2.$$

$$NiS + 2O_2 = NiSO_4.$$

At approximately 900°C.

$$NiSO_4 = NiO + SO_3.$$

Both these latter reactions occur with a great liberation of heat which can cause fusion of the catalyst mass, and if the reactions are taking place in a section of the bed not covered by a thermocouple, the condition can proceed undetected for some time. However, it is suggested that this sulphating occurs chiefly with catalysts which have even distribution of nickel in the particles. As shown earlier, the magnesite catalyst has its nickel concentrated on the outer surface.

The mechanical maintenance required has not been excessive, and in comparison with normal c.w.g. plants the balance is in favour of the Onia Gegi plant; this is to be expected because there are no gritty solids to handle and no mechanical grate to maintain. When using medium fuel oil the stack valve received most attention, mainly removing carbon from the valve seat.

The maintenance of mechanical and electrical equipment has been in accordance with good engineering practice, and consequently little gas-making time has been lost due to breakdowns. Some anxiety was caused, however, by the corrosive effluent problem, when shutdowns of several hours duration became necessary for renewal or patching of corroded steelwork on the scrubbers and effluent handling flumes. Major repairs were avoided until the plants were shut down for the annual inspection.

Both gas-making units were shut down for inspection after one year. The refractory linings to combustion chambers and base of pre-heaters had suffered some slight spalling and were hard burnt, consistent with that expected from the combustion of fuel oil. The top layers of chequers below the vaporising chambers were in good condition, although under the make-oil sprayers some disintegration had taken place and ash had been deposited between the chequer brick squares. A considerable amount of ash was removed from the galleries at the top of the reactors, but no restriction had been caused.

The linings of the reactors were in good condition but, as with the pre-heater steam inlet ports, areas around the tertiary burner ports were spalled by action of steam and water. The arches supporting the catalyst beds and chequers had several open joints, and in one case a slight settlement had occurred, otherwise they were sound.

Cracking of the reactor, combustion chamber and pre-heater linings had taken place, probably during the shut down by contraction on cooling, but they were of only a minor nature. Erosion was observed on the outlet of the reactor of one unit where the hot gases changed direction before entering the waste heat boiler.

Flame impingement

The top catalyst beds on both units had suffered flame impingement from the secondary burners and on one unit the top surface was broken down and fused together. This led to the introduction of a protective layer three inches deep consisting of refractory bricks broken down to 1½ in. cubes. The bottom catalysts in both units were in good condition but, as with the top beds, had ash deposited on them. One unit had a complete change of catalyst from magnesite to sillimanite ready for reforming H_2S-free refinery gas, while the other was left untouched.

The only major work required on the gas washing and cooling plant, other than that brought about by the corrosive effluent and wash water, was the renewal of the primary scrubber wooden packing and the removal of part of the secondary scrubber packing which had fallen into the base of the scrubber. The failure of the scrubber packing was partly attributed to the large quantity of naphthalene produced during catalytic gasification of medium fuel oil and in the period when reforming refinery gas with a high H_2S content. The latter necessitated working with high catalyst temperatures to increase catalyst activity, and thus favoured the formation of naphthalene.

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Technical developments

A new pipe bender

A COMPLETELY new type of pipe bender, to be known as the 'field' bender, has been especially designed as a lightweight two-stage hydraulic unit, primarily for site work in oil fields and construction projects. The substitution of the orthodox heavy beam or winghead by hinged arms, which fold into a neat compact unit, enables the machine to be made extremely portable for easy transportation.

Supplied in two sizes, one with a range of formers for bending pipes from 2-in. to 4-in. bore and the other for bending pipe from 4-in. to 8-in. bore, it consists of a Staffa heavy duty hydraulic ram to the base of which is attached the oil reservoir and arm trunnion brackets, arms carrying the pipe clamps which are telescopic and simply positioned by means of a pin, and a weatherproof pump unit mounted on top of the tank.

The Chamberlain hydraulic pipe bender

The formers supplied enable pipe of 2-in., 2½-in., 3-in. and 4-in. nominal bore to be bent in one operation up to an angle of 90°, without heating or filling and without removal from the former. With pipes above these sizes, the radii are produced by successive bending.

Power-driven units may be supplied—air/hydraulic pumps for site work, when compressed air is available, and electric motors for shop use.

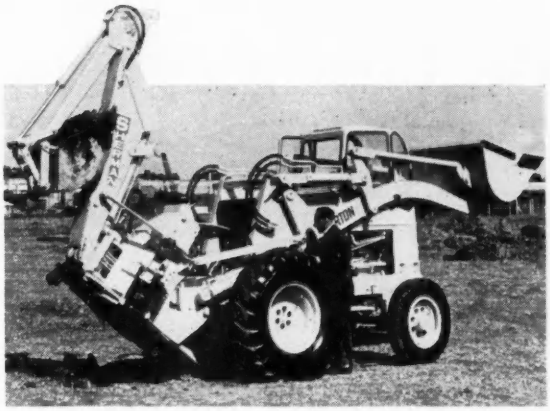
Larger machines, mechanically or hydraulically operated, and capable of bending up to 18-in. diameter pipe, are available.—*The Chamberlain Group of Companies.*

New digger loader

THE Merton Sherman Universal is the latest addition to the Merton range of loading shovels and allied equipment. Combining the hydraulic Frontloader 59 and the Sherman Panther power digger, it embodies advanced design and operating features. Unlike the usual long, clumsy loader/digger arrangement, the Universal is a compact and well-balanced unit, only 17-ft. long in the 'loading shovel' or 'carry' position, with a performance, manoeuvrability and stability completely unaffected by the additional equipment it carries.

In addition to 'semi-automatic' hydraulic controls which take over routine operations and speed the loading cycle, the front-mounted driver's cab gives the operator a perfect, unobstructed view of the bucket and the ground in front.

The raising and lowering of the Sherman Panther digger can be carried out by one man using a remote controlled lever suitably placed on the outside of the machine. It is a robust and fast power digger with its hydraulic stabilisers individually controlled at the touch of a lever, allowing the machine to be set up level on sloping or uneven ground with its hydraulic cylinders well protected from damage.



The Merton Sherman Universal digger loader

Operating under a pressure of 2,000 p.s.i. with a special multiple operating control valve, the Sherman digger has a very fast operating cycle of 15 to 19 seconds and its speed of digging, in the hands of an experienced operator may be between 25 and 75 cu. yards per hour, depending on bucket size and site conditions.

The power unit is a Fordson Major 4-cylinder diesel of 54-b.h.p. at 1,800 r.p.m.; a constant mesh gear box gives six forward and two reverse speeds.

The Frontloader 59 is constructed on a heavy rolled steel channel main frame with a robust fabricated front axle and centre pivoted. The buckets available vary from ¾ cu. yard to 2 cu. yards capacity and rock buckets, bull- and angledozer blades, fork lift and crane attachments can be supplied.—*Merton Engineering Co. Ltd.*

New refractories

THE increasingly severe conditions of modern blast furnace operation demand new and better refractories. G.R. has met this demand by improvements in the existing Foster range and the introduction of completely new types of blast furnace refractories.

Among these are the Numax, a new dense, high fired refractory to meet the most exacting conditions encountered in the stack in modern furnace operations; the Sillmax 'I,' a 100% bondless Sillimanite refractory for use in the lower stack, bosh and hearth, and the Foster high grade 'S' which is an improved quality stack refractory with a high abrasion resistance and resistant to carbon monoxide.

Also in the series are the Foster Crown 'S,' an improved quality 42%-alumina refractory; the Foster Carbon, a well-tried and proved refractory for use in hearth, well and bosh, and the complete range of G.R. Amberlite insulating material. Suitable jointing cements and ramming compositions are available for these various refractories.—*General Refractories Ltd.*

Dry materials hose

GOODYEAR has added to its range of industrial hoses a lightweight, flexible hose specially built for the bulk handling of dry materials which are not highly abrasive. It is particularly well suited for the handling of grain, dry chemicals in crystal powder form, fertilisers and similar materials.

The Goodyear dry materials hose (Style 273 × 243) is supplied in lengths to suit the site and application. It can be built with enlarged ends for fitting couplings or with soft ends for clipping to existing pipework or hopper feed tubes. The seamless $\frac{3}{16}$-in. smooth-bore tube of high grade rubber is reinforced by three plies of strong, closely-woven fabric and a helix of round steel wire embedded in the rubber, while the cover consists of a medium-gauge, abrasion-resisting compound. The hose has a maximum working pressure of 50 p.s.i. and is supplied in bore sizes of 2½ in. and 4 in.—*Goodyear Tyre & Rubber Co. (Gt. Britain) Ltd.*

Oblique valve

THE new Winn air-operated oblique valve which efficiently damps out concussion when closing, is designed for controlling water, steam and other suitable media at pressures up to 100 p.s.i. Internally, the valve incorporates a clean, straight-through design which reduces the possibility of trapping solids or foreign matter in the region of the valve seat. Apart from the ease of maintaining the general cleanness of the bore, frictional resistance is reduced while the speed of flow is increased. The valve is opened by air at 40 p.s.i. line pressure and automatically closed by spring action on release of the air pressure. The range of sizes of this new Winn valve is from ½ in. to 1½ in. in bronze and 2 in. in cast iron, the latter being available in quantity from stock.—*Winn & Coles Ltd.*

A Winn air-operated oblique valve

Flameproof diesel tractor

ITS compact design allows the MT.25 to work where no other tractor could operate; its powerful, reliable diesel engine, four-wheel drive on large diameter tyres and simplicity of control give maximum performance under all conditions. Rugged construction and ease of access to all working parts ensure the minimum of maintenance. The water-cooled exhaust system and Hunslet patent 'jet' type exhaust conditioner ensure maximum safety and minimum air pollution. The extensive range of ancillary equipment allows the basic tractor to be used for a wide variety of applications.—*The Hunslet Engine Co. Ltd.*

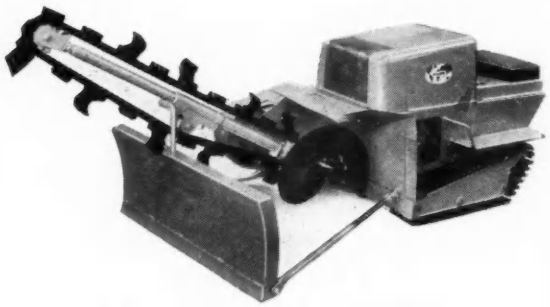
New small trencher and backfiller

THE Allen 3/12 is a small, versatile trencher and backfiller, ideal for builders, plumbers, councils, public works contractors, telephone, water, gas, and electrical undertakings, etc.

The Allen 3/12 can dig trenches 3 in. to 4 in. wide, down to 5 ft. 6 in. deep, or 12 in. wide at 2 ft. 6 in. depth, with a range of intermediate widths and depths. The digging boom is controlled hydraulically, and can be telescoped to suit the depth of trench.

A specially-designed backfiller blade is available for the Allen 3/12, which can be set straight, or angled to the left, by a simple pin setting. Blade raising and lowering is hydraulically controlled.

The power unit is a 12½ h.p. four-stroke engine, air-cooled, with 12-volt electric starter. A built-in tool box is provided.



The Allen trencher and backfiller

All controls are grouped within easy reach of the operator from his padded seat. The drive mechanism is a one-piece unit, with hardened gears and heavy-duty bearings used throughout. A two-speed reduction unit gives trenching speeds from 1-12 ft. per minute, with a high speed of up to 2 m.p.h. for travelling or bulldozing.

Independent drive clutches, and disc steering brakes are incorporated for each track, with instant forward and reverse by hydraulic power from valve on control panel. Weight distribution on the tracks gives a ground pressure of only 3.3 lb. per sq.in., enabling the machine to negotiate soft going without sinking or bogging.

The digging chain is V-belt driven, and engaged by spring-loaded idler in the on- and off-positions. The teeth are made from heat-treated alloy steel. Combinations of teeth and teeth-brackets enables trenches of all widths from 3 in. to 12 in. to be dug. A pre-loaded shock absorber device is incorporated in the boom, which cushions the shocks encountered while digging without loss of power.

Spoil is deposited from the digging boom on to an augur, which rotates with a spiral motion to discharge on the left-hand side of the machine. A sweep blade is mounted on the right, to leave a clean surface adjacent to the trench.

Leading dimensions of the Allen 3/12 are:—

Height (ground level to top of hood) ..	3 ft. 1 in.
Width, overall	3 ft. 1 in.
Length (to front of auger)	6 ft. 4½ in.
Weight	1,350 lb.

For transportation of the machine from site to site, a specially-designed trailer is available, with tilting platform for easy on and off loading.—*John Allen & Sons (Oxford) Ltd.*

JUNE H.P. BUSINESS DOWN ONE-FIFTH ON LAST YEAR BUT—

Only gas appliance sales seriously affected are cookers

APART from gas cookers, sales of domestic gas appliances have not been much affected by the hire purchase restrictions, although Board of Trade figures for June published last week showed H.P. business in general to be one-fifth down compared with the same period last year.

This is the more remarkable since durable goods shops were the most affected and electricity and gas showrooms are included in this category by the Board of Trade.

NEWS SUPPLEMENT

A leading appliance manufacturer told the 'GAS JOURNAL' that although cooker sales were down 20% on 1959, H.P. restrictions were not the only factor since sales last year benefitted from the lifting of restrictions. Also, the drop in cooker sales was balanced by an increase in refrigerator sales.

June deliveries of British-made refrigerators to the home market were up 18.2% on 1959 and January-June deliveries were up 45% compared with the corresponding figure for 1959.

Cooker sales down

A London gas board sales manager admitted that the drop in cooker sales was some 10-15% greater than was expected when restrictions were first re-imposed. Sales were, however, running at approximately the 1955-56-57 level and were expected to rise now that people could see that the increase in H.P. rates was not so bad. Refrigerator and central heating sales he stated, were going quite well.

Last week's Board of Trade figures showed that the decline in sales was accompanied by a slower rise in total H.P. debt. During June the debt increased by only £3 mill. to a total of £697 mill., after increasing by £29 mill. in April and £15 mill. in May.

Sales down in June

The *Financial Times* of June 9 said that sales of durable goods during June were estimated to have declined by 10%. A leader on the same day pointed out that controls had had no effect on the amount the public spent on hire purchase but simply meant that larger down payments bought smaller quantities of goods.

'There is,' the *Financial Times* commented, 'a basic clash between national stability and the stability of individual industries, and until some better method is devised the consumer goods industries seem liable to suffer.'

GAS IN THE AIR

A PLANE drawing a huge 'Gas Cuts Costs' banner is backing up an intensive central heating advertising campaign being run by the North Western Gas Board.

The plane is flying over all the centres of population in the Manchester-Liverpool industrial belt, as well as in the North and West Lancashire and Cheshire areas.

£15,000 has been spent by the Board on press advertising alone during the last two weeks of June and the month of July. The 100 flying hours booked will cost an additional £2,000.

N.E.G.B. Club takes to water

THE 140-strong North Eastern Gas Board Sports and Social Club at Scarborough, formed two years ago, have gone in for boating and fishing. Last December they bought a lifeboat from a Hull trawler and it has been converted so that it can carry about 20 people on sea cruises or parties of ten on fishing trips.

PERSONAL NOTES

MR. A. C. MURDOCH has been appointed Chief Technical Officer of Teddington Refrigeration Controls Ltd. as from August 8. He will have jurisdiction over all engineering fields, including development and project engineering of products for refrigeration and heating control. Mr. Murdoch was Chief Research and Development Engineer to the Prestcold Division of the Pressed Steel Co. Ltd. He is an examiner in Refrigeration Practice for the City and Guilds Institute, London.

MR. D. F. HAYDON has been appointed General Sales Manager of Baird & Tatlock (London) Ltd. Mr. Haydon joined the company in 1932, and gained experience in the various commercial departments of the organisation before becoming a Technical Sales Representative in the early part of 1939. In 1952, he became the Chief Technical Sales Representative.

ADVICE SERVICE ON INDUSTRIAL EFFLUENT

SIMON-CARVES LTD. and Monsanto Chemicals Ltd. have pooled their experience in the treatment and disposal of industrial effluent to form the Simon-Carves Monsanto Effluent Advisory Service, it was announced last week. The service is to be administered by Simon-Carves Ltd. from their headquarters at Cheadle Heath, Stockport, Cheshire.

Advice on effluent

At home and overseas the Simon-Carves Monsanto Effluent Advisory Service will provide general advice on effluent problems, on-site and laboratory investigations on the treatment of specific effluents, recommendations and design data for effluent treatment processes and plants, and continuing advice on the operation of commissioned treatment plants.

Available for construction

The engineering and contracting organisation of Simon-Carves Ltd. is available for the construction, installation and commissioning of complete effluent treatment plants.

Simon-Carves Ltd. have carried out in recent years, and are still engaged on, intensive laboratory, pilot-plant and field-scale research and development on the treatment of effluents by biological methods.

They have eight commercial-scale biological effluent treatment plants in service, under construction and on order at colliery and steelworks coking plants in the United Kingdom and India, and are dealing with enquiries for the treatment of effluents from many other industrial processes.

OBITUARY

MR. HARRY S. DAVIES, B.E.M., Mains Superintendent, South Wales Gas Grid, died suddenly on July 22, aged 56. Mr. Davies served nearly 34 years with the gas industry in South Wales. He joined the Rhymney & Aber Gas Company in 1926 and was concerned mainly in the distribution of coke oven gas to towns in the Rhymney Valley area of South Wales. When the Wales Gas Board commenced the construction of the Eastern Section of the South Wales Gas Grid in 1951, Mr. Davies was appointed Mains Superintendent and continued in this capacity until the time of his death. In 1955 he was awarded the B.E.M. for his services to the gas industry.

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